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MgO Filtration Research

By D. N. Tallman and J. E. Pahlman



UNITED STATES DEPARTMENT OF THE INTERIOR

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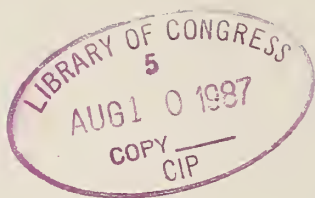
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UNITED STATES DEPARTMENT OF THE INTERIOR
Donald Paul Hodel, Secretary

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UNIT OF MEASURE ABBREVIATIONS USED IN THIS REPORT

°C	degree Celsius	L/s	liter per second
cal/(mol·deg)	calorie per mol per degree	m	meter
cm	centimeter	mg/L	milligram per liter
cm ³	cubic centimeter	min	minute
cm H ₂ O	centimeter water (pressure)	mm	millimeter
cm ³ /g	cubic centimeter per gram	μm	micrometer
cm/s	centimeter per second	μmho	micromho
eV	electron volt	m/s ²	meter per second per second
g	gram	m ² /s	square meter per second
g/cm ³	gram per cubic centimeter	mV	millivolt
h	hour	NTU	nephelometric turbidity unit
K	kelvin	pct	percent
kg/m ³	kilogram per cubic meter	ppm	part per million
kg/(m·s)	kilogram per meter per second	r/min	revolutions per minute
L	liter	s	second
L/cm ²	liter per square centimeter	vol pct	volume percent
L/min	liter per minute	wt pct	weight percent
		yr	year

MgO FILTRATION RESEARCH

By D. N. Tallman¹ and J. E. Pahlman²

ABSTRACT

The Bureau of Mines has completed 4 yr of deep-bed filtration research comparing the efficiencies of granular magnesium oxide (MgO) and conventional filter sand in single- and dual-medium filters when filtering mineral-processing and mine waters. Bench and field evaluation tests were conducted using water types ranging from synthetic suspensions of asbestos, kaolin, metal hydroxides, and milled sand to process waters from magnetite beneficiation, iron ore flotation, granite cutting, and a steel mill. Even though the turbidity reductions were similar, MgO filters were found to be advantageous when filtering water pretreated with alum because the volume throughput before breakthrough (turbidity > 1 NTU unit) was much more for the MgO filter than for the sand filter. Reduced head loss owing to the greater porosity of the MgO filter beds is potentially the most beneficial advantage in employing MgO as the filter medium, especially when filtering water pretreated with polymer flocculants. Granular MgO (periclase) is durable and can be backwashed like sand. It is compatible with anthracite in a dual-medium filter and apparently is not poisoned by dissolved metals in the process water. No single solid-removal mechanism could be identified for the improved filtration observed with granular MgO.

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INTRODUCTION

Suspended solids are a common impurity in mine water. Mines rely mainly on flocculation and settling for solid removal, but this can be unreliable owing to seasonal variation in water temperature, runoff, and changes in the character of ores being mined. Filtration is often needed to provide sufficiently pure water to meet mineral processing requirements and/or statutory effluent limits. Although recycling decreases the volume of discharge needing treatment, it can adversely affect plant performance unless the return water is treated to keep contaminants from reaching unacceptable levels of concentration.

The Bureau of Mines has completed 4 yr of research on improving filtration of mine and mineral-processing water using MgO as the filter medium in contact and deep-bed filters. The initial phase of work resulted from a previous study of the surface charge of asbestos fibers in water (1).³ Numerous candidate materials (sand, calcite, diatomaceous earth, acidic alumina, basic alumina, microcrystalline cellulose, magnesium carbonate, and activated carbon) were tested for removing asbestos fibers from water by contact filtration, a process that required no pretreatment of the water and that used shallow beds of fine filter media. MgO filter media gave the best removal. The superior performance of MgO was attributed to its positive surface charge (2). The next phase of research was the application of MgO to filtration of other suspended solids occurring in natural waters. Bench-filtration tests that

compared MgO and filter sand were performed on synthetic suspensions and on mine-water samples (2-3). In these tests granular materials were used for practical reasons such as achieving adequate solid-loading capacity and reducing pressure drop through the filters. Flocculation with alum improved filtration with the granular media, and MgO generally outperformed sand. Field tests were run to validate bench-scale tests on mine water (4).

The remaining phase of research dealt with practical engineering aspects of using MgO filters, such as fluidization, backwashing, and durability. Attempts to derive a model for MgO bed expansion were unsuccessful because there appears to be a transition in the expansion behavior of the MgO for particles between 0.5 and 1.0 mm. MgO was found to be compatible with anthracite but not with sand in dual-medium filters as the anthracite and MgO were easily restratified by backwashing. Granular MgO possesses the necessary durability to be a filter medium and was apparently not poisoned by dissolved metals in the process water. The MgO filters were tolerant to moderate levels of calcium hardness and carbonate alkalinity, provided adequate backwashing with air scour was available. Cementation of MgO grains with each other or with anthracite grains was observed with scale formation in the filters. Descriptions of each phase of research and discussion of pertinent results are presented in this report.

BACKGROUND

FILTRATION MECHANISMS

A detailed theoretical discussion of filtration mechanisms as they pertain to MgO filtration is beyond the scope of this paper. Reviews of the theory of deep-bed filtration are available in the

literature (5-7). A brief examination of relevant filtration mechanisms, however, is helpful for understanding the filtration results presented herein.

In deep-bed filtration, filter pores are usually large compared with the size of the particles being filtered. Clearly, forces other than straining must account for particle retention (5). Transport and attachment steps are both important in the capture of small

³Underlined numbers in parentheses refer to items in the list of references preceding the appendix.

suspended particles. Flow is usually laminar, so forces must act on a particle to move it across streamlines into close proximity with the surface of the filter medium where attachment forces operate.

The main forces considered in particle transport are gravitational, diffusional, hydrodynamic, and inertial forces (4-5). Surface forces that affect attachment are of mainly two kinds, the molecular dispersion force (London van-der-Waals force) and the double-ionic-layer force, which is mainly due to surface charge (6). Particle interception is really neither a transport mechanism nor an attractive force, but it is the final step in particle capture in any case. The net effect of all these forces will determine the overall removal efficiency of the filter. Contributions to particle collection by each type of force can be estimated using the dimensionless parameters given in table 1. In general, it can be seen that the size of the particle

to be captured is the single most important variable; increasing the particle size will enhance all collection mechanisms except diffusion and London van-der-Waals forces. For particles smaller than 1 μm , surface charge, ionic strength, and diffusion are dominant capture mechanisms, while interception and/or straining becomes dominant above 10 to 100 μm . For particles with sizes between 1 and 10 μm , mixed capture mechanisms are observed. Overall collector efficiency is at a minimum for particles of about 1- μm diam; increasing particle size by flocculation results in better collection. Flocculation also acts to decrease surface charge on the particle so that surface charge effects are reduced. At high ionic strength, the ionic-double-layer thickness ($1/\kappa$) is minimal. This also reduces surface-charge effects.

Particle detachment is especially important in cleaning filters, but also

TABLE 1. - Dimensionless filtration parameters

Parameter	Typical range	Formula	Comments
ATTACHMENT MECHANISMS			
$N_{DL} \dots\dots$	10^0-10^5	κa_p , where $\kappa = \left[\frac{4\pi e^2}{\epsilon_d kT} \sum_j m_j z_j^2 \right]^{1/2}$	Dimensionless double-layer-thickness effect of ionic strength versus conductivity.
$N_{E1} \dots\dots$	$10^{-1}-10^3$	$\epsilon_d \kappa (\psi_p^2 + \psi_m^2) / 12\pi \mu v$	Electrokinetic group 1: electrical versus shear forces.
$N_{E2} \dots\dots$	-1 to +1	$2\psi_p \psi_m / (\psi_p^2 + \psi_m^2)$	Electrokinetic group 2: "parity" group; - = favorable, + = unfavorable interaction.
$N_{LO} \dots\dots$	$10^{-5}-10^{-2}$	$h / 9\pi \mu a_p^2 v$	London group: London-van-der-Waals forces-shear forces.
INTERCEPTION AND STRAINING			
$N_R \dots\dots$	$10^{-4}-10^{-1}$	d_p / d_m	Relative-size group.
TRANSPORT MECHANISMS			
$N_G \dots\dots$	0 -14	$g a_p^2 (\rho_p - \rho_f) / 18\mu v$	Gravity group: settling velocity-approach velocity, dimensionless.
$1/P_e \dots\dots$	$10^{-8}-10^{-5}$	$\frac{RT}{3\pi \mu d_p d_m}$	Inverse of Peclet number: ratio of diffusion number to approach velocity.
$Re \dots\dots$	~ 1	$vd_p g / \mu = vd_p / v$	Reynolds number: characterizes hydrodynamic; usually laminar.

Sources: Rajagopulan and Tien (7) and Ives (5).

during filtration, where increases in flow rate will create proportionately higher shear forces that can dislodge deposited solids and cause breakthrough. Beds with large filter-medium grains and large porosity will have less head loss and will be more resistant to scouring and subsequent breakthrough (4).

TRANSIENT BEHAVIOR

To get a complete description of a filtration process, one needs to obtain a record of the pressure drop (both across the bed and within the bed), filtrate quality, and flow rate as they vary with time. Idealized versions of typical kinds of head-loss and turbidity behavior are illustrated in figure 1 by curves *A* and *A'* for deep-bed filtration and by curves *B* and *B'* for surface straining or cake filtration. Filtration is a dynamic process with pressure and turbidity fluctuating as channels become clogged with deposits and reopened by scouring. As filtration progresses, more material becomes lodged in the filter, the porosity decreases, and as a result head loss increases with time. This manifests itself as a rise in the water level above a gravity filter or as increased gauge pressure in a pressure filter. At constant loading, the deposit in the filter grows at a steady rate, and the head above the filter increases almost linearly with time (curve *A*, fig. 1). Changes in either influent flow rate or solids concentration will have a corresponding effect on head loss. This is the expected pattern in deep-bed filtration. Often the suspended solids are actually removed at the surface of the bed by straining. As a result of this phenomenon, a cake forms and pressure begins to increase exponentially, as in curve *B* of figure 1. Filtrate turbidity will generally be very low, as in curve *B'*. Surface or cake filtration is inefficient, because designed increases in allowable head loss result in diminishing gains in filtrate volume before backwashing is necessary. Poor design, changes in solid loading, or attrition of the medium

are the usual causes of excessive rates of head loss.

In deep-bed filtration the filtrate turbidity initially decreases slightly and then levels out before breakthrough (curve *A'* of figure 1). Up to a point, material deposited in the filter improves solid capture, but then a critical point is reached where the reduction in porosity and the resulting increased shear rates cause a decrease in the efficiency of solid removal. In fact, deposits are often dislodged and then trapped deeper in the filter. This scour-deposit mechanism was postulated by Mints (5) when he observed that particles appearing in the filtrate at breakthrough were larger than those in the influent, leading to the assumption that they were actually flocs of deposited material. Onset of breakthrough may manifest itself as increased fluctuation in the filtrate turbidity and is most easily observed with a turbidimeter-recorder setup. In Bureau filtration tests, increased fluctuation in the turbidity was usually observed prior to consistent increases in turbidity. This may be a reliable indicator of imminent breakthrough.

FILTRATION INDICES

Efficient deep-bed filtration is the result of a successful compromise between filtrate quality goals and minimizing head loss. The best filter achieves the desired quality at the lowest pumping cost. Flocculant dosage and mixing, filter-grain size, flow rate, and depth are all interdependent; sometimes adjustments in one or more of these variables become necessary due to variations in influent composition. Numerous attempts have been made to describe the performance of a filter quantitatively, in terms of both filtrate quality and head loss, with a single number or index (5, 8). One of the simpler and more commonly used indices is the filterability index (*F*) given by Ives (5):

$$F = (C/C_0)(H/vt) \cong (\tau/\tau_0)(H/V), \quad (1)$$

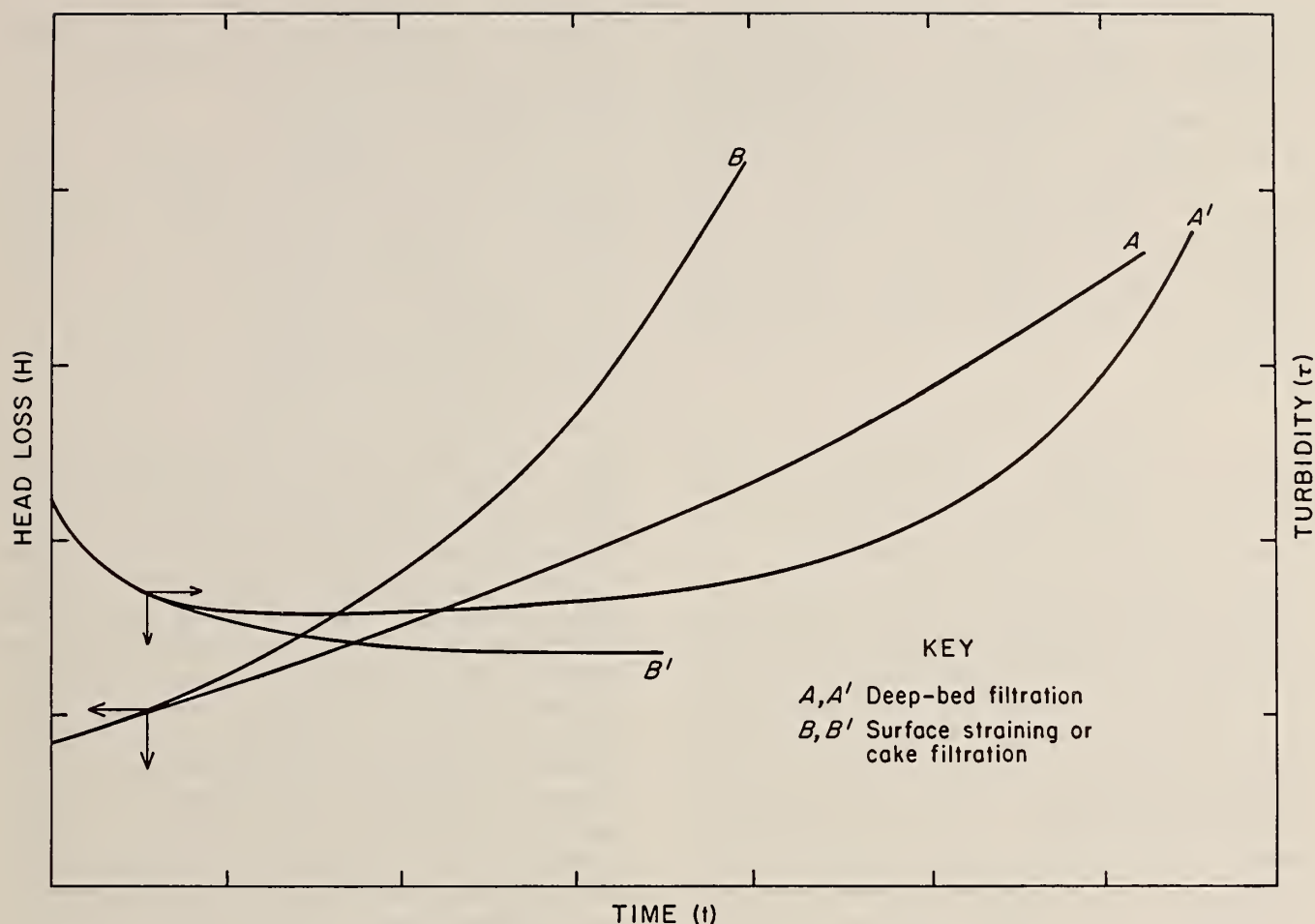


FIGURE 1.—Transient behavior typical of granular bed filtration.

where C = concentration,

H = increase in head loss,

v = filtration rate or approach velocity,

t = time,

τ = filtrate turbidity,

V = filtrate volume per unit area at time t ($V = vt$),

C_0 = influent concentration,

and τ_0 = influent turbidity.

The filterability index is the product of two dimensionless parameters that can be represented by residual turbidity and head-loss rate. Low values of F indicate

efficient filtration; effective solid removal and/or small head loss could contribute to a low value for F . F could be used, for example, by a plant operator who needs to adjust plant-water treatment to seasonal variations of influent-solid loading. Several options such as flocculant dosage and flow-rate changes would be tested on several small-scale filters run in parallel. The parameters that result in the minimum value of F and still meet quality goals can then be tried full scale. Comparisons of filter media can also be made with F ; the filter giving the lowest value of F is the best choice for a given influent. Other indices are available that are similar to F but take into account operating limitations on head loss and filtrate quality for a particular plant (8).

Higher solids concentrations will increase H/V , but this is not accounted for by F . Comparisons of tests run with

different solids concentrations require a different index that will normalize the effect of concentration. The solid capture index (SC) measures the amount of solid trapped in the filter per unit area per unit head loss and is expressed in units of concentration (g/cm^3). This type of index was used by Bauman and Cleasby (8) to compare waste water filtration results at various treatment plants and is defined by the following equation:

$$\begin{aligned} \text{SC} &= C_o(1-C/C_o)(V/H) \\ &= C_o(1-\tau/\tau_o)(V/H). \end{aligned} \quad (2)$$

Large values of SC are desirable.

One disadvantage of using filtration indices is that different values of H, V, and τ could give the same value for F or SC. For the large number of tests performed in this study, filtration indices provide the only practical means of analyzing experimental data. Aside from a few figures used to illustrate special cases, the filterability index, F, and solid capture index, SC, have been used throughout this review. To offset the ambiguity that results from relying on filtration indices, the values of τ/τ_o and V/H are given so that the contribution of each parameter can be assessed.

CONTACT FILTRATION OF ASBESTOS FIBERS

Contact-filtration experiments were performed using a number of natural and synthetic filter materials. The filters were typically a 0.5- to 2.0-cm layer of 0.10-mm (minus 100- plus 200-mesh) Baker⁴ reagent-grade material supported by a layer of coarse sand and cotton. MgO, sand, magnesium carbonate, acidic and basic alumina, calcite, diatomaceous earth, microcrystalline cellulose, and activated carbon were all tested against suspensions of amosite and crocidolite asbestos in distilled water. Amphibole and chrysotile asbestos samples were obtained from the International Union

This is especially important when values of either F or SC for two filters are almost equal.

The values of V and v are also listed for further evaluation of results. Generally, values of F were calculated for the portion of a test preceding breakthrough. After breakthrough, F values may actually improve because often there is a reduction in head loss that is disproportionate to average turbidity increase. Comparison between a filter that has broken through and one that has not broken through is invalid because the former no longer meets quality requirements. Excellent values of F (due to low head loss) may be obtained for a filter, but breakthrough may be reached so quickly that it would be impractical to use that filter. A ratio of V/v greater than 21.6 (run length equal to 6 h) should be achieved for practical purposes; net production of filtered water declines rapidly with further decreases in run length because more frequent backwashing becomes necessary (9). In some cases, filtrate quality never reaches satisfactory levels or may be declining when a run is terminated because of excessive head loss on one of the filters. For these tests, F values are based on the entire run, and values of V for the two media being compared should be approximately equal.

Against Cancer (UICC). Ultrasonic and mechanical agitation were used to disperse the asbestos fibers in water. The beds were washed with 50 cm^3 of distilled water, 15 cm^3 of 1- to 10-ppm suspension was added to the next 50- cm^3 aliquot passing through the filter, and finally the filter was washed with an additional 50 cm^3 of water. Flow velocities were 0.07 to 0.15 cm/s under 10- to 20-cm gravity head. The entire volume of filtrate was then passed through a 0.45- μm -pore-size membrane filter. Fiber counts were made on sections of filter membrane using a scanning electron microscope at magnifications of $\times 1,000$ to $\times 5,000$.

Results for the contact filtration experiments are summarized in table 2. MgO

⁴Reference to specific products does not imply endorsement by the Bureau of Mines.

TABLE 2. - Contact filtration of asbestos,
percent removed

Solid filter medium	Amphibole	Chrysotile
Sand.....	0	0
Calcite.....	10	60
Diatomaceous earth.....	60	10
Microcrystalline cellulose....	20	20
Alumina, acidic.....	99	50
Alumina, basic.....	50	80
MgO.....	100	99
Carbon.....	85	65

Source: Schiller and Khalafalla (2).

and acidic alumina were both efficient for filtering amphibole asbestos, but only MgO could effectively remove both amphibole and chrysotile asbestos. These results can be explained on the basis of surface charge and pH. Surface charge on particles in water is pH dependent. Charge is acquired owing to selective dissolution-adsorption of ions at the liquid-solid interface. Amphibole asbestos and most naturally occurring particulates are negatively charged in water. Adsorption of H^+ and OH^- account for the pH dependence; particles generally become more negative at higher pH. The isoelectric point for amphibole asbestos is pH 3 to 3.5, and that of chrysotile is

approximately pH 11. Amphibole asbestos has a negative zeta potential of -25 to -45 mV in neutral and slightly basic water. Chrysotile has a positive zeta potential but flocculates in water at high pH. MgO is basic and has a positive surface charge below pH 12, so it is able to remove the amphibole asbestos fibers by electrokinetic attraction and still remove positively charged chrysotile by causing it to flocculate. Acidic alumina is also positively charged, but not basic, so it collected amphibole asbestos but passed the chrysotile because the pH of the suspension was not increased enough to cause flocculation.

GRANULAR BED FILTRATION

The contact filters restricted flow excessively and collected too few particles to be directly applicable for treating water. For practical reasons such as solid loading, filtration rate, and backwashing (filter regeneration), granular filters are used to filter waste water.

The purpose of this phase of work was to compare red flint sand and MgO as granular filters for removal of asbestos and other naturally occurring particulates. Granular MgO was purchased as crushed periclase from Basic Chemical and Kaiser. The periclase is made by roasting MgO (calcining) at a high enough temperature so that it fuses and becomes inert ("dead burning"). This product was chosen because it was less friable than pellets of a more active MgO material. Active magnesia such as was used in the

contact filtration studies is calcined at lower temperatures so that much of its internal porosity is retained. Consequently, in changing to a durable granular material substantially more surface area was lost than would be predicted on the basis of gross-particle size.

Filtrate quality is often measured by turbidity (light scattering) rather than particle counting because this allows real-time monitoring and better control of the filtration process. Filtrate turbidity was therefore used as a measure of filtering efficiency in almost all of the granular-bed-filtration tests. Initially, a comparison was made between particle counting with the scanning electron microscope and turbidity measurement (nephelometric), and the two methods were found to agree within experimental error of the former. Pressure and turbidity

were measured at 15-min intervals over runs lasting several hours or more.

It became apparent that granular MgO was not nearly as effective for removal of asbestos as was the more active MgO used in contact filtration. Adequate removal could be achieved only by adding flocculant to the influent. Figure 2 illustrates how removal rates of both sand and MgO filters are improved by the addition of 1 ppm of alum (potassium aluminum sulfate heptahydrate) to the suspension 15 min before filtration. Flow velocity was 0.5 cm/s. These results indicate that forces other than surface charge operate in these filters. Although the MgO gives better removal than sand in the absence of alum, both filters benefit by the increase in particle size due to flocculation. For the 0.71-mm sand and MgO used in this test, bed porosities are 38 and 50 pct, respectively. With flocculation the sand actually becomes the more efficient filter by a small margin,

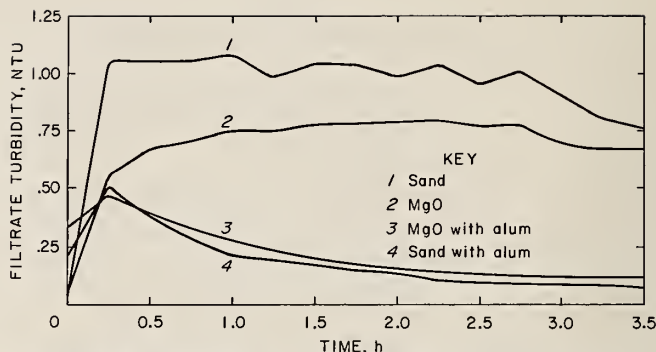


FIGURE 2.—Filtration of amphibole asbestos by granular MgO and sand filters with and without alum pretreatment.

owing to enhanced particle interception or mechanical straining. However, the increase in pressure drop (head loss) across the sand filter occurs from two to five times faster than that for the MgO filter. In virtually all remaining tests flocculant was added before filtration.

SYNTHETIC SUSPENSIONS

Standard suspensions of kaolin and milled sand were used in many filtration tests to achieve maximum reproducibility of solution turbidity. Mine water samples, in contrast to standard suspensions, are often complex mixtures of various dissolved chemicals and particulates that may vary considerably between sampling and testing. With consistent preparation procedures, synthetic suspensions of known concentrations and reproducible turbidities were made. In-mine-water sample turbidity was assumed to be proportional to suspended-solid concentration, and actual suspended-solid concentrations were not measured for every test. For the synthetic suspensions, it was possible to calculate both filterability (F) and solid-capture (SC) indexes because suspended-solid concentrations were known. Only F was determined for mine-water samples, but the effect of altered influent-solid loadings can be estimated from head-sample turbidities. The mass of flocculant was not included in the calculation of SC , even though aluminum hydroxides could contribute to solid loading as a result of using alum for flocculation.

FILTRATION OF KAOLIN

Results for the filtration of kaolin suspensions by 30-cm-deep beds of MgO and sand are listed in table 3. Single-medium MgO filters are better than sand filters by a substantial margin; the value of F is about eight to nine times smaller and that of SC is about four times larger for MgO than for sand. Comparison of tests 1 and 2 demonstrates how the effect of suspended-solid concentration is minimized by using SC rather than F . The higher suspended-solid concentration causes a larger rate of head loss and consequently increased the value of F , but this effect is factored out in SC . Using finer sized MgO decreases the efficiency of the filter; filtrate quality is no better with 0.42-mm MgO, and head loss rate is approximately doubled. Since shear rates are considerably higher in the 0.42-mm MgO, run length was shortened by earlier breakthrough.

Dual-medium filters were made with a 15-cm layer of 0.71-mm sand over 30 cm of either 0.42-mm MgO or 0.42-mm garnet sand. Garnet sand is sometimes used as a second or even third layer in

TABLE 3. - Filtration of synthetic suspensions

Test	Filter medium (d _m)	Flocculant, mg/L	Suspended solids, mg/L	τ/τ_0	V, L/cm ²	v, cm/s	V/H	F ($\times 10^5$)	SC, g/cm ³
KAOLIN									
1.....	0.71-mm sand.....	5	24	0.025	4.03	0.22	17	140	0.40
2.....	...do.....	5	5	.125	11.7	.46	140	87	.63
3.....	0.71-mm MgO.....	5	5	.047	13.4	.46	480	9.9	2.20
4.....	0.42-mm MgO.....	5	5	.069	8.3	.46	260	27	1.30
5.....	...do.....	5	5	.060	8.2	.46	106	57	.50
6.....	0.71-mm sand, 0.42-mm garnet.....	5	5	.038	7.5	.35	107	36	.51
7.....	...do.....	10	25	.023	7.5	.35	79	29	1.9
8.....	0.71-mm sand, 0.42-mm MgO.....	5	5	.044	7.2	.33	260	17	1.2
9.....	...do.....	10	25	.011	7.3	.34	116	9.2	2.9
10.....	1.3-mm anthracite, 0.42-mm MgO.....	25	200	.106	3.6	.25	47	230	8.4
11.....	1.3-mm anthracite, 0.42-mm garnet.	25	200	.179	3.3	.23	44	400	7.3
MILLED SAND									
12.....	0.71-mm MgO.....	2.5	7.3	0.066	11.6	0.46	830	8.0	5.6
13.....	0.71-mm sand.....	2.5	7.3	.099	12.5	.46	125	85	.76
14.....	0.71-mm sand, 0.42-mm garnet.....	15	160	.009	5.5	.45	98	8.8	16
15.....	...do.....	15	160	.005	2.9	.41	55	9.2	8.7
16.....	0.71-mm sand, 0.42-mm MgO.....	15	160	.004	3.0	.41	71	5.5	11.4
17.....	...do.....	15	160	.007	7.1	.42	169	4.2	27
18.....	...do.....	15	160	.005	5.9	.41	84	6.4	13
19.....	...do.....	15	160	.006	6.3	.41	98	5.5	18

multiple-medium filters. The performance of the two dual-medium filters is more nearly equal than that of the single-medium filter. Since in both filters the sand (upper) layer removes most of the suspended solids and consequently contributes most to head loss, the filters would be expected to be similar. The sand-MgO filter is 1.5 to 3 times more efficient than the sand-garnet filter.

Dual-medium filters of anthracite-MgO and anthracite-garnet were tested with much higher loadings of kaolin. These filters were comprised of 50 pct anthracite and either 50 pct MgO or 50 pct garnet for an overall depth of 30 cm. Although run lengths were cut in half, a large amount of solids were collected by these filters. Filtering such high loadings of kaolin is probably impractical because filters would be operating at <80-pct availability because of the need for frequent backwashing.

FILTRATION OF MILLED SAND

MgO filters were dramatically better than sand filters for filtering milled sand flocculated with alum (table 3). MgO gives lower F values by a factor of 10, and SC values are seven to eight times larger than for sand. Both filters were 42 ± 1 cm deep. With dual-medium filters the difference is much less apparent. The dual-medium filters were 46 cm deep, with the top third of the filter being the sand layer. $F(\times 10^5)$ ranges from 8.8 to 9.2 for sand-garnet filters and from 4.2 to 6.4 for sand-MgO filters. Although the variance between values for duplicate tests is fairly large, the differences between filter-medium types are statistically significant. The mean values are 9.0 for $F(\times 10^5)$ and 12.4 g/cm^3 for SC for the sand-garnet filter versus 5.4 and 17.4 g/cm^3 for the sand-MgO filter. Filtrate volume at breakthrough for the sand-garnet filter is about 60 pct that of the sand-MgO filter. The sand-MgO filter is 1.4 to 1.6 times better by any of these criteria.

There is remarkable similarity between results for the filtration of kaolin and milled sand in these tests. Values for SC and F are better overall for filtering

the milled sand. This is probably due to particle size differences between the two suspensions and to the use of higher loadings of milled sand. It is easier to get a high removal rate with increased influent loading. The relative differences between the various filters are quite consistent for the two suspensions.

USE OF POLYMER FLOCCULANTS

Kaolin Filtration

Recent trends in process-water treatment include the use of organic polymers or polyelectrolytes to improve flocculation and increase filter-loading capacity. Often an inorganic salt of aluminum (alum) or Fe^{3+} is added to destabilize colloidal dispersions and create small "pin" flocs. Then a polymer is added to further flocculate the solid into large (1- to 10-mm), fast-settling flocs. Typical dosages of polymer are 0.1 to 10 ppm, and the amount required is strongly dependent on the solids concentration. Although most of the solids are removed effectively by settling, a small residual fraction of solids often has to be removed by filtration. The choices of polymer (cationic, anionic, or nonionic) and dosage are usually determined by settling tests. Flocs created with polymer are larger and more shear resistant than alum-type flocs. Coarser media, deeper beds, and higher filtration velocities are typical of filtration with polymer flocculation, and solid-loading capacity is usually larger than with alum flocculation.

Sand-MgO and sand-garnet filters were tested with a 25-ppm kaolin suspension treated with 10 ppm alum and 1.0 ppm Separan AP-30, an anionic polymer; the results are given in table 4. The best filtration was with the sand-garnet filter and a filtration velocity of 0.35 cm/s. The coarser anthracite-sand filter gives approximately equal SC, but residual turbidity is relatively high: $\tau/\tau_0 = 0.22$. The best result for MgO was with coarser media and a velocity of 0.52 cm/s, but in general it can be seen that use of polymer actually diminished

TABLE 4. - Filtration with polymer flocculants

Test	Filter medium (d _m)	Flocculant, mg/L		τ _o , NTU	τ/τ _o	V, L/cm ²	v, cm/s	V/H	F (×10 ⁵)
		Alum	Separan AP-30						
FILTRATION OF KAOLIN									
20.....	0.71-mm sand, 0.42-mm MgO.....	10	1.0	27	0.040	9.3	0.34	18	220
21.....	0.71-mm sand, 0.42-mm garnet.....	10	1.0	27	.022	10.6	.35	68	32
22.....	1.3-mm anthracite, 0.71-mm MgO.....	10	1.0	27	.075	9.4	.52	14	520
23.....	1.3-mm anthracite, 0.71-mm sand.....	10	1.0	29	.22	14.8	.51	100	210
24.....	1.3-mm anthracite, 1.3-mm sand.....	10	1.0	27.5	.26	14.8	.51	22	1,150
25.....	1.3-mm anthracite, 1.3-mm MgO.....	10	1.0	27	.11	11.2	.52	26	410
HEAVY METAL PRECIPITATES									
26.....	0.71-mm MgO.....	NA	NA	18	0.80	3.1	0.20	180	430
27.....	0.71-mm sand.....	NA	NA	18	.71	3.1	.20	170	410
28.....	0.71-mm MgO.....	NA	NA	7.0	.15	2.5	.35	180	84
29.....	0.71-mm sand.....	NA	NA	7.0	.11	2.5	.35	100	110
	0.71-mm MgO.....	NA	0.2	17	.019	3.0	.41	5.9	320
	0.71-mm sand.....	NA	.2	17	.020	1.4	.40	2.6	760
30.....	1.3-mm MgO.....	NA	.2	17.5	.022	7.3	.41	16	140
	1.3-mm sand.....	NA	.2	17.5	.023	5.7	.39	10	230
31.....	1.0-mm MgO.....	NA	.2	13.5	.020	5.7	.40	10	200
	1.0-mm sand.....	NA	.2	13.5	.023	5.0	.38	9.6	240
32.....	1.8-mm MgO.....	NA	.2	15	.055	2.6	.41	53	100
	1.8-mm sand.....	NA	.2	15	.045	2.6	.41	53	100
33.....	0.71-mm MgO.....	NA	NA	3.2	.085	5.9	.41	13	650
	0.71-mm sand.....	NA	NA	3.2	.058	3.0	.42	6.0	970
34.....	0.71-mm MgO.....	NA	NA	3.4	.048	10.7	.50	19	254
	0.71-mm sand.....	NA	NA	3.4	.032	9.9	.46	17	190
35.....	1.7-mm MgO.....	NA	NA	3.8	.062	13.5	.47	190	33
	1.7-mm sand.....	NA	NA	3.8	.064	14.1	.49	120	54

NA Not available.

the filterability of the MgO. Filtration indexes for the sand-garnet filter are about the same whether polymer or only alum was used to flocculate the kaolin. Head loss through the sand-MgO filter was greatly increased; this occurred almost entirely in the MgO (lower) layer. Usually the upper layer collects the greater share of solids and constricts flow more than the lower (finer) layer, and the sand and garnet contribute little to the observed head loss when they comprise the lower layer. The filterability index $F(\times 10^5)$ for MgO was increased from 9.2 with alum flocculation (table 3, test 9) to 220 by using anionic polymer (table 4, test 20). Solid capture of the sand-MgO filters decreased by a factor of 21. Under these conditions, MgO did not offer any significant advantage over conventional media.

Filtration of Metal Precipitates

Fe^{3+} and other base-metal ions are commonly found in mine water. Precipitates of these ions as hydrous metal oxides present difficult filtration problems because of their fragility; they are easily disrupted by the shear forces encountered in conventional granular-bed filters. These precipitates also settle slowly, and the voluminous toxic sludge produced by settling presents a serious disposal problem. Polymer is often added to improve settling and to aid in dewatering.

A series of tests were performed to see if MgO offered any advantage over conventional filter media for postsettling filtration of metal hydroxides. Sand and MgO filters were first tested on $\text{Fe}(\text{OH})_3$ precipitates, produced by bringing a 100-ppm- FeCl_3 solution to pH 8.7. A second test suspension was prepared that had 78 ppm Fe^{3+} and 10 ppm Mn^{2+} as sulfate salts and 37 ppm Mn^{7+} as permanganate at pH 8.0 to 9.0. The Mn^{7+} was present at about 20 pct excess for oxidizing the Fe^{2+} and Mn^{2+} , so the predominant species under these conditions were MnO_2 and $\text{Fe}(\text{OH})_3$. A similar suspension was made by simply oxidizing 11 ppm Fe^{2+} with 1.7 ppm $\text{Mn}(\text{VII})$ at pH 7.5 to 8.5. Separan AP-30, an anionic polymer, was

used to flocculate the precipitates in most of the tests. These suspensions were made to simulate waste water from a western metal mine that contained Mn and Fe at approximately these levels. Field tests were later run on the actual mine effluent. In the mine water treatment plant, polymer was used to aid clarification and also for sludge dewatering; no additional polymer was added to the water being pumped through the small test filters, since none was being added to that entering the sand filters in service at the plant.

Results for both laboratory and field tests are summarized in table 4. Bed depths for all tests were 30 cm. Neither medium could filter prepared suspensions of metal oxides effectively without the use of polymer. Field tests and laboratory tests with flocculant were in reasonable agreement, especially if the reduced head-sample turbidity in the field tests is taken into account. During the field tests plant water was cleaner than normal because the mill operation had been suspended and the water being treated was diluted by spring runoff. Reduced solid loading would result in F being lower than predicted from laboratory tests. The most efficient filtration (minimum F) is with 1.7- and 1.8-mm media, and the MgO is slightly better than equivalently sized sand. In general, sand gives slightly better removal rates and the MgO has smaller head losses, probably owing to the larger porosity of the MgO. Surface-charge forces would be expected to diminish in importance when the filter media and the suspended-particle size are large, and because flocculation has also essentially neutralized particle surface charge. The stronger and larger flocs produced by the polymer are retained in the filter mainly by gravity or hydrodynamic forces, so the two media are essentially equally effective. Flocs were able to penetrate well into both filters, so straining was not the dominant removal mechanism. Filtering of the base metal hydroxides or hydrous oxides, even with the use of polymer, is fairly inefficient compared with filtering particulates such as kaolin.

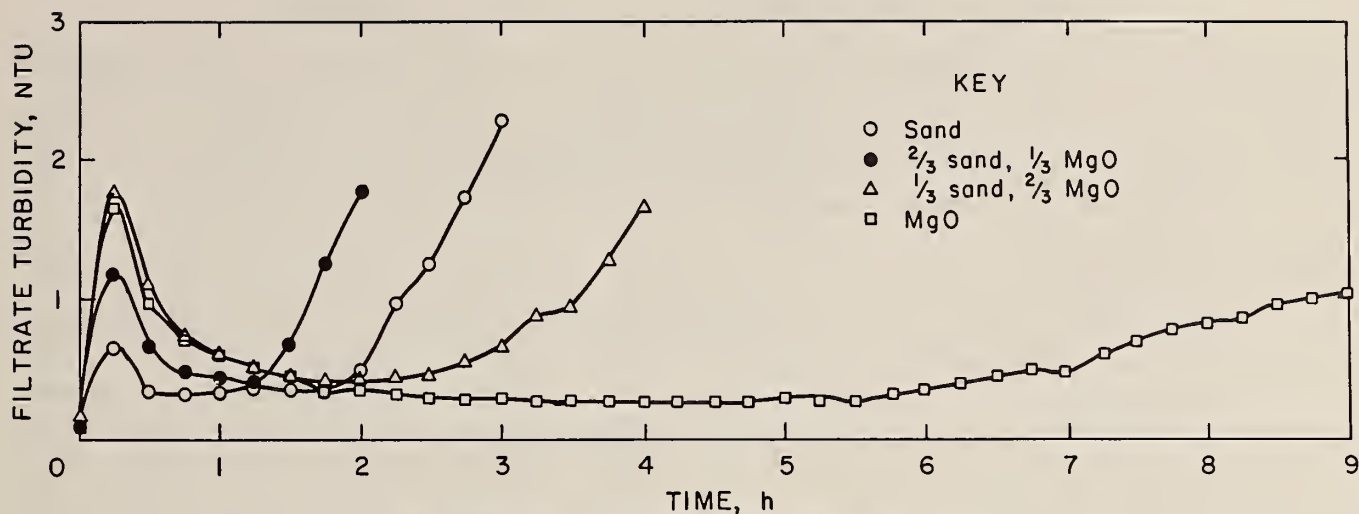


FIGURE 3.—Filtration with mixed MgO-sand filters.

Efficiency might be improved by using higher filtration rates and deeper beds of coarser media.

Filtration With Mixed MgO-Sand Beds

A brief investigation of the possibility of using mixed MgO-sand beds was undertaken. However, as shown in figure 3, this does not appear to be feasible because of rapid breakthrough with mixed beds. The filters were 30-cm beds of

0.71-mm MgO and/or sand, and the filtration velocity was 0.21 cm/s. Because of the poor retention of solids, head loss decreased with decreasing MgO fraction. Values of SC were 4.8, 5.2, 8.3, and 9.6 g/cm³ in order of increasing sand content, but in this case the use of indexes is misleading. The most likely reason for the poor retention of kaolin by the mixed beds is the reduction in porosity that results from mixing spherical (sand) and angular (MgO) particles.

APPLICATIONS

MgO and sand filters were compared in field tests at four mines and two mineral-processing sites. In addition to mine water, MgO was tested on river water for its application to municipal-water treatment.

STEEL MILL COOLING WATER

Large volumes of process water are used in secondary production of steel. Much of this water is used in the direct-spray quenching of billet and becomes contaminated mainly by mill scale and tramp oils. The mill scale is removed by settling primary and secondary scale pits, but residual suspended solids clog sprayer nozzles and increase abrasion of plant piping. Tramp oils are treated by skimming and emulsification with surfactants.

The treated water passes through a cooling tower before returning to the plant.

Several samples of return water were filtered in laboratory tests before field testing began. Results for both field and laboratory tests are shown in table 5. The process water had high levels of dissolved solids, and suspended particles had weakly negative zeta potentials, but these could not be measured accurately owing to the high conductivity of the water. Jar tests showed that 30 ppm alum gave optimum settling. Results of the bench tests indicated that the dual-medium filters were not significantly better than single-medium filters, and that the MgO and anthracite-MgO filters were able to filter two to three times more water before breakthrough than the sand or anthracite-garnet filters.

TABLE 5. - Filtration of steel mill cooling water

Test	Filter medium (d _m)	τ_0 , NTU	τ/τ_0	V, L/cm ²	v, cm/s	V/H	F ($\times 10^5$)
LABORATORY TESTS							
36.....	1.3-mm anthracite, 0.71-mm sand...	13.3	0.038	3.4	0.34	106	36
37.....	1.3-mm anthracite, 0.71-mm MgO....	19.3	.018	6.7	.34	91	20
38.....	0.71-mm MgO.....	24	.015	7.4	.34	106	14
39.....	0.71-mm sand.....	27	.030	2.4	.34	57	53
40.....	0.71-mm MgO.....	22	.011	6.2	.34	31	35
FIELD TESTS							
41.....	0.71-mm MgO.....	32	0.015	3.38	0.35	13	120
	0.71-mm sand.....	32	.056	1.84	.34	3.1	1,800
42.....	0.71-mm MgO.....	33.5	.023	3.40	.33	18	130
	0.71-mm sand.....	33.5	.054	1.50	.33	8.7	620
43.....	0.71-mm MgO.....	25.3	.018	8.35	.21	38	47
	0.71-mm sand.....	25.3	.022	2.93	.21	42	53
44.....	0.71-mm MgO.....	23.5	.025	>10.8	.18	>59	<42
	0.71-mm sand.....	23.5	.025	5.17	.15	33	76
45.....	0.71-mm MgO.....	19	.019	5.54	.23	30	63
	0.71-mm sand.....	19	.018	2.61	.23	31	57
46.....	1.3-mm anthracite, 0.71-mm MgO....	18	.021	2.32	.24	41	52
	1.3-mm anthracite, 0.71-mm sand...	18	.024	1.38	.24	39	63

Values of $F(\times 10^5)$ are 14 and 35 for MgO and 53 for sand, while differences between the dual-medium filters are less pronounced.

In the field mainly single-medium filters were tested, and flow rate was the only variable that was manipulated. In general, the field and laboratory tests are in agreement. Filtration velocities had to be reduced from 0.34 to 0.21 cm/s to get F values roughly equivalent to those from bench tests, but this appears to be mainly due to differences in head loss. Higher suspended-solid loading, entrained air (possibly enhanced by use of surfactant for oil breaking), and other factors could be responsible. Field tests were usually controlled to a lesser degree and generally gave slightly poorer results than laboratory tests. Filtrate volume at breakthrough for the MgO filter often doubled that for the sand filter, and F values were lower except for one test. Differences between dual-medium filters probably were not statistically significant, and run lengths were shortened considerably by early breakthrough. Although the field test results were for the most part not quite so good (higher F) as results from

laboratory tests, the trends in the data are remarkably similar, and MgO is substantially more effective than sand in this application.

PROCESS WATER FROM MAGNETITE BENEFICIATION

Process water used in the beneficiation of magnetite becomes laden with moderate levels of suspended hematite. Water is recycled and is treated only by settling before returning to the grinding circuit. As can be seen from the results of bench tests given in table 6, this water is easily filtered when 15 ppm alum is added for flocculation. MgO filters give slightly higher F values but more than double the filtrate volume of sand at breakthrough. Dual-medium anthracite-garnet and anthracite-MgO filters are nearly equivalent. Three field tests were run with single-medium MgO and sand filters, and the differences were even more pronounced than in the laboratory tests; breakthrough occurred so rapidly in the sand filters that quantitative comparisons were meaningless. Results for the MgO filters were comparable with those from bench tests.

TABLE 6. - Filtration of iron ore processing water

Test	Filter medium (d_m)	τ_o , NTU	τ/τ_o	V, L/cm ²	v, cm/s	V/H	F ($\times 10^5$)
47.....	1.3-mm anthracite, 0.71-mm MgO....	49.5	0.005	4.0	0.34	220	2.4
48.....	1.3-mm anthracite, 0.71-mm garnet.	46.0	.003	4.3	.34	120	2.2
49.....	1.3-mm anthracite, 0.42-mm MgO....	54.5	.003	8.3	.33	990	.26
50.....	0.71-mm sand.....	55.0	.003	4.3	.34	51	5.0
51.....	0.71-mm MgO.....	56.0	.002	11.2	.34	21	9.3

PROCESS WATER FROM FLOTATION OF IRON ORE

Process water from this flotation operation had almost 250 ppm of very fine iron oxide and silicate particles, 11 ppm Ca hardness, 110 ppm dissolved SiO_2 , 115 ppm P alkalinity, and 610 ppm M alkalinity (as milligrams of CaCO_3). This water is first settled in a large sedimentation basin and then clarified with alum flocculation before discharge. In these tests overflow from the sedimentation basin was used to test the MgO and sand filters.

Filtration results were extremely sensitive to flocculant dosage (Separan AP-30 anionic polymer), which was varied between 0.62 and 2.5 ppm. In general, sand outperformed MgO in field tests, giving F values that were 2.5 to 5 times lower than those for MgO. $F(\times 10^5)$ ranged from 470 to 1,300 for MgO and from 200 to 290 for sand. These values are too high (high rate of head loss) to be a practical application for either filter medium. The results are consistent with other tests in which MgO filters became plugged easily when polymer flocculants were used. This type of process water needs to be softened and clarified before attempting filtration.

MISSISSIPPI RIVER WATER

Mississippi River water is used to supply potable water to a large segment of the Minneapolis-St. Paul metropolitan area. Municipal-water treatment involves lime-softening, settling, and filtration followed by disinfection. A series of filtration tests were run on Mississippi River water treated in a manner similar to that used at the municipal water treatment plant. River water was first softened with 150 ppm CaO and then

flocculated with 2.0 ppm alum. The decant after 15 to 30 min settling was at pH 8.0 ± 0.5 and had turbidity of 35 ± 2 NTU. More alum was mixed with the decant so that final alum concentration in the filter influent was 26 ppm. Thirty-centimeter beds of sand and MgO were tested in parallel at a filtration velocity of 0.21 cm/s.

Three bench-scale tests were run under these conditions, and in two of the three tests run lengths were extended enough to reach breakthrough. Values of $F(\times 10^5)$ averaged 13 ± 4 for the MgO and 22 ± 12 for the sand. The main difference appears to be in the lower rate of head loss for the MgO, which averaged 1.7 times lower than for sand. The MgO filtered about 28 pct more water at breakthrough, based on the average of two tests.

Since the minimum run length was already about 11 h, both filters would operate at >95 pct availability, so the extra filtrate volume per cycle is probably not too significant. The reduced head loss rate is an important advantage, however.

PROCESS WATER FOR CUTTING GRANITE

The final field tests were conducted on water used in cutting granite. Particulate-laden water is collected from all cutting and grinding stations and settled in a clarifier. Polymer is used to aid flocculation, but clarification is poor owing to inadequate control of flocculant dosage, inadequate mixing, and convection due to evolution of gases caused by microbial activity. On occasion phosphoric acid (H_3PO_4) is also added to the system when a special kind of finish is being applied to the granite. The process water is recycled with minimal makeup water; consequently,

hardness and conductivity are increased and calcite scaling and corrosion are serious operating problems. Zeta potentials of the suspended particles were found to be weakly negative, with values ranging from -1.4 to -23 mV with a mean value of -11 ± 6 mV. The specific conductivity was $1,300 \mu\text{mho}$.

These tests were intended to measure the filtering capability of the MgO over an extended period of time. Media attrition and scaling effects were to be observed directly, and their impact on filter performance evaluated. Laboratory tests on process water samples indicated that MgO, sand, and anthracite were approximately equal in their ability to filter this water. Both alum flocculation and softening the water by elevating pH were more effective than using polymeric flocculants.

In the field tests the water was treated first by softening with 6 pct NaOH and settling, and then by filtration through 55-cm beds of either 0.71-mm MgO or sand. The NaOH solution was metered into raw process water entering a 570-L settler, and the overflow was kept at

pH 9.0 to 9.8. Total flow through the system was 11.4 L/min (0.19 L/s), and the filtration velocity was 0.31 cm/s . Dual-medium filters were also tested with 1.3-mm anthracite replacing the upper 25 to 30 pct of the beds.

Laboratory tests varied considerably but generally indicated that $F(\times 10^5)$ values of around 30 could be expected in the field tests. Unfortunately, field test results varied even more than the laboratory tests. In some cases turbidity removal was excellent for both filters, and in other tests neither filter would perform efficiently. No trend could be discerned that would indicate that attrition was impairing MgO performance or that either sand or MgO was the better filter. Figure 4 illustrates the wide variations in filtrate quality for tests with dual-medium filters. Curve 3 is a fairly typical plot for a good filtration run, in this case with the anthracite-MgO filter. Filtrate turbidity decreases below 1.0 NTU within 1 h and remains low for a reasonable length of time. The anthracite-sand filter (curve 4) also achieved

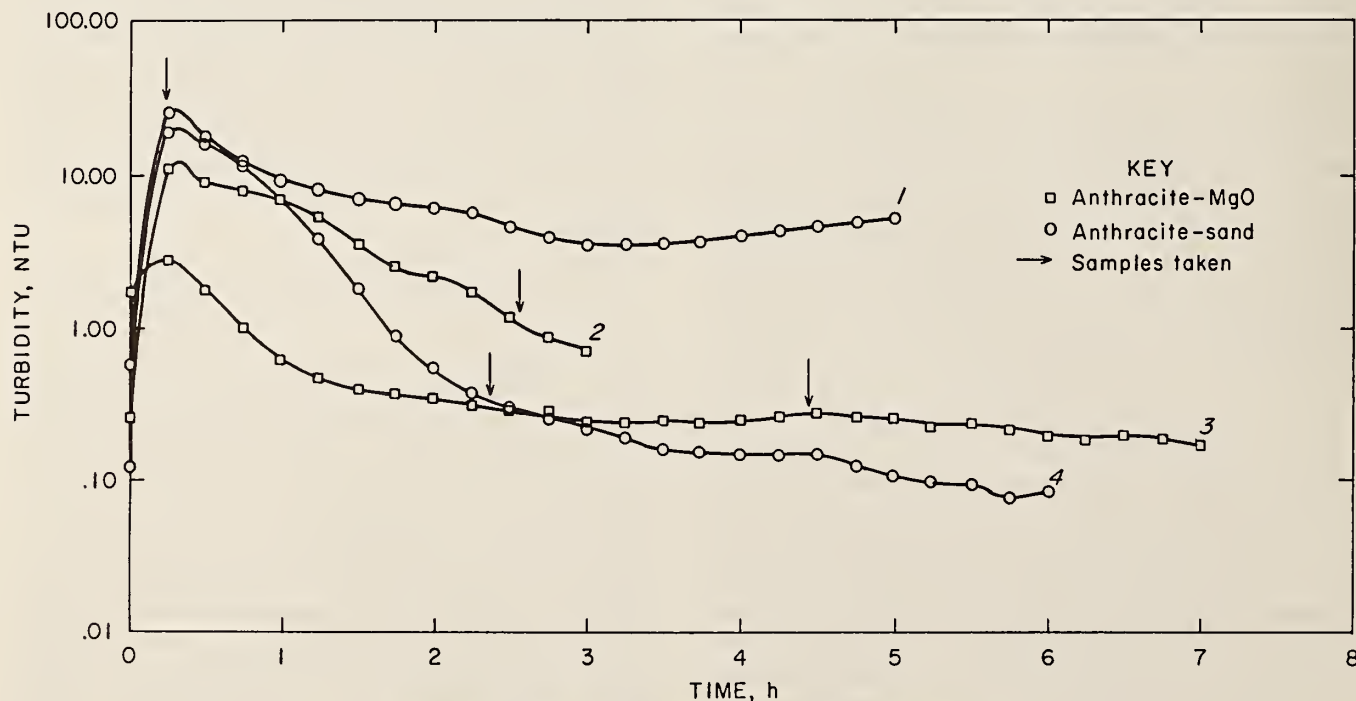


FIGURE 4.—Filtration of granite-cutting process water.

excellent turbidity removal after a longer delay.

The arrows in figure 4 correspond to sampling for chemical analysis at the times shown. From the results given in table 7, it appears that good filtration resulted when P (probably as orthophosphate) and Fe both were present in the process water. On the day that test 1 was run both P and Fe levels were low and the anthracite-sand filtrate was turbid. After 5 h flow was switched to the anthracite-MgO filter, and turbidity removal was poor in the initial stages. Toward the end of the run, filtrate removal was improved and P and Fe were again both present at measurable levels. It is noteworthy that the MgO removes P and the sand does not. The MgO also adds small but significant amounts of Ca and Mg to that in the influent. (CaO is

an impurity in the periclase, present at 1 to 3 wt pct.)

TABLE 7. - Chemical analysis of process water

Stream	pH	Concentration, mg/L				
		Ca	Mg	S	P	Fe
Test 1:						
Influent..	9.3	77	57	54	<1.0	1.7
Filtrate..	9.3	68	55	56	<1.0	<.2
Test 2:						
Influent..	9.3	71	55	60	1.6	3.0
Filtrate..	9.4	79	66	60	<1.0	<.2
Test 3:						
Influent..	9.4	64	64	69	1.3	2.3
Filtrate..	9.5	70	71	65	<1.0	<.2
Test 4:						
Influent..	9.3	69	63	66	1.7	.24
Filtrate..	9.3	65	62	56	1.3	3.1

ENGINEERING ASPECTS

BACKWASHING OF MgO FILTERS

Filters are cleaned by backwashing with an upward flow of water, which fluidizes the bed and removes deposited solids mainly by hydraulic shearing forces. Often air scouring is used to increase the backwash efficiency and reduce both the flow rate and flow volume required to clean the filter. The air bubbles increase abrasion between filter grains and give high shear rates owing to the increased turbulence in the bed. After dual-medium filters are backwashed with air scouring, a high-rate backwash is required to restratify the media mixed by the vigorous cleaning action. Understanding and being able to predict bed expansion are important for both the design and operation of deep-bed filters. Ideally, it would be possible to calculate bed expansion for any flow velocity knowing only the basic properties of the filter medium and the fluid. In practice bed expansion can only be predicted from correlations derived from experimental data. It is also important to predict optimal backwash rates and to determine whether a desired combination of filter

materials can be restratified by backwashing when used together in dual-medium filters.

A filter bed becomes fluidized when the flow velocity is increased to the point where the head loss across the medium is equal to the weight of the grains in water. Further increases in flow velocity do not increase head loss across the bed; the bed expands and the increase in porosity offsets the higher velocity. A mass balance on the medium requires that

$$h_b (1-\epsilon) = h_{b0} (1-\epsilon_0), \quad (3)$$

where h_b = bed depth,

ϵ = porosity,

h_{b0} = unfluidized bed depth,

and ϵ_0 = unfluidized bed porosity.

Bed porosity is observed to increase logarithmically as a function of flow velocity. In dimensionless form this is expressed as

$$\log Re = n \log \epsilon + \log Re_i, \quad (4)$$

where Re = Reynolds number,
 $Re = (v d_m / \nu)$,

ν = the fluid velocity,

d_m = the filter medium grain size,

ν = the kinematic viscosity of the fluid,

and Re_1 = the Reynolds number corresponding to the unhindered settling velocity of the filter grains.

As this velocity is approached, the medium is entrained in the fluid and the porosity becomes unity. Maximum shear rates occur at porosities between 0.68 and 0.71 (10).

The parameters n and Re_1 are obtained by measuring the initial-bed porosity or bulk density; for an arbitrarily chosen starting depth of medium, expanded bed height is measured at several flow velocities. Expanded-bed porosity is calculated from equation 3, and linear regression or graphical methods are used to find the slope and intercept of plots of $\log \epsilon$ versus $\log v$. This is repeated for various sizes of each medium.

Both n and Re_1 are dependent on the fundamental dimensionless group known as the Archimedes or Galileo number:

$$Ar = \frac{g d_m^3 [(S.G.)_m - (S.G.)_f]}{\nu^2}, \quad (5)$$

where Ar = ratio of buoyant to viscous drag forces acting on a particle,

$(S.G.)_f$ = specific gravity of the fluid,

$(S.G.)_m$ = specific gravity of the medium,

g = the acceleration owing to gravity (9.8 m/s^2),

d_m = the diameter of the grains of medium,

and ν = the kinematic velocity (m^2/s).

Ar incorporates only properties intrinsic to the medium and fluid. Values of Ar range from 10^2 to 10^5 for conventional deep-bed filters.

Normal backwashing spans a transition region between laminar and turbulent flow, with Re_1 ranging from less than 10 to approximately 500. One approach to modeling bed expansion is to use power-log correlations between n , Re_1 , and Ar for each medium:

$$\log n = a_1 \log Re_1 + b_1 \quad (a_1 < 0) \quad (6)$$

$$\log Re_1 = a_2 \log Ar + b_2. \quad (7)$$

This is successful if done within sufficiently narrow limits of Ar . For laminar or Stokes settling ($Re < 1$), this simplifies to

$$Re_1 = (\text{const}) Ar, \quad (8)$$

and for fully turbulent flow ($Re > 100$), a_1 becomes very small and n approaches a constant value. Although accurate models can be obtained in this manner for a specific filter medium, attempts to make a more general model fail because there is no meaningful measure of particle shape and its effect on drag coefficient (10-11).

A simple correlation was given by Bohm (12) for predicting the flow rate corresponding to maximum mass transfer rate in a fluidized bed:

$$Re_{opt} = 0.072 Ar^{0.614}. \quad (9)$$

High shear rates enhance mass transfer by reducing the boundary layer thickness around the particles. Because large shear rates are also needed for cleaning deep-bed filters, this correlation may be useful for predicting optimum backwashing velocities.

Fluidization curves for six sizes of MgO are shown in figure 5. Similar curves were obtained for sand and

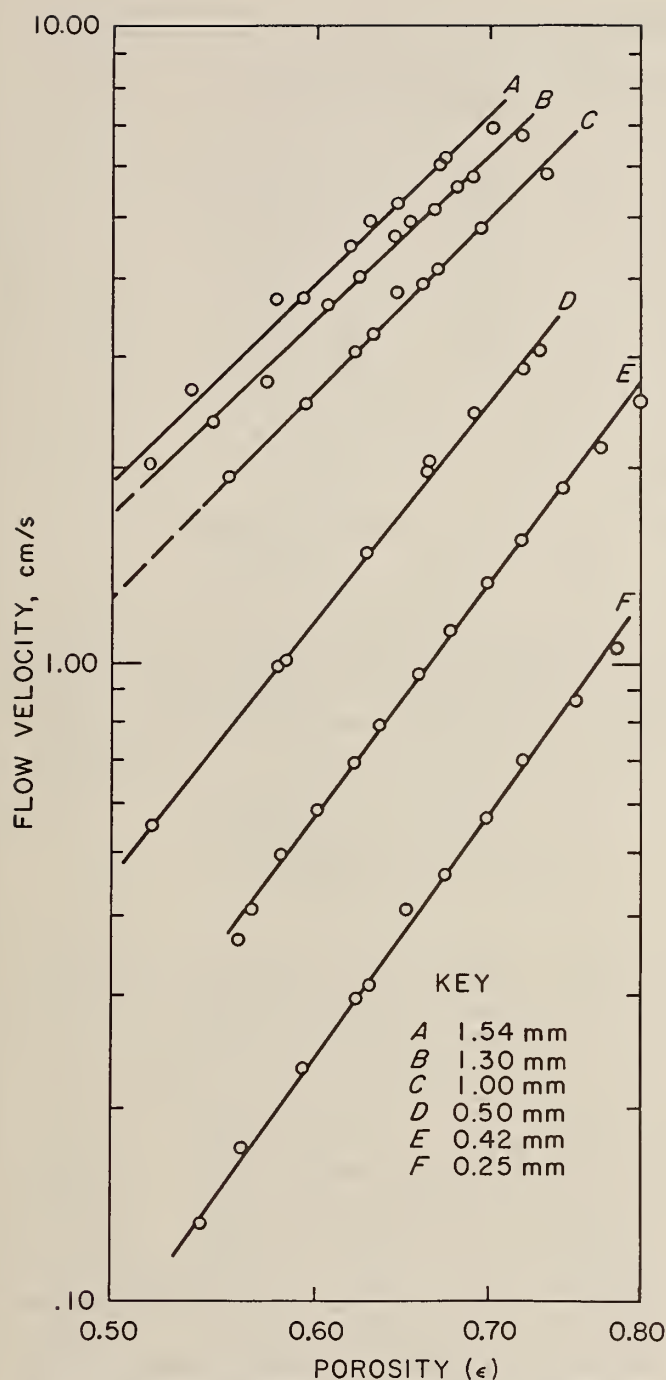


FIGURE 5.—Bed expansion of MgO versus flow velocity.

olivine. Values of n and Re_i were determined by linear regression of the $\log \epsilon$ versus $\log v$ plots. Literature values for the viscosity and density of water were used to calculate Ar and Re at the temperatures measured during each test.

Test results are summarized in table 8. Also included are values given by Gunasingham (10) for anthracite, ballotini, polystyrene, and sand. Re_{opt} was calculated from equation 9 and then substituted for Re in equation 4 to get ϵ_{opt} .

Attempts to derive a specific model for the expansion of MgO were unsuccessful owing to the nonlinearity of plots of $\log n$ versus $\log Re_i$ and $\log Re_i$ versus $\log Ar$. At grain sizes of 1 mm and greater, n varies much less than at the smaller sizes. Apparently there is a transition in the expansion behavior of MgO particles between 0.5 and 1.0 mm. Data for sand also displayed nonlinear characteristics, but values for n agree fairly well with those of Gunasingham (10), considering probable differences in particle shape and size distribution not taken into account. Because of these uncertainties, this type of correlation seems of limited value for design purposes.

Calculated optimum bed porosities fall between 0.56 and 0.68 for particle sizes greater than 0.5 mm (fig. 6). Neglecting the lower ϵ_{opt} porosity values for ballotini and sand, the values for MgO, polystyrene, and anthracite are between 0.65 and 0.68, which is in good agreement with the predicted porosity range needed to produce maximum hydraulic shear (10). Ballotini and sand have smaller optimum porosities because they are more spherical and have smaller initial porosities than MgO, anthracite, and polystyrene. Below 0.5-mm grain size the optimum bed porosity is substantially greater than the predicted range for maximum shear. This is consistent with the data given by Bohm (12), which showed that, for $Ar < 10^3$, Re_{opt} values begin to lie well below the correlation given by equation 9. For nonspherical particles larger than 0.5 mm, the simpler correlation successfully predicts the optimum backwash velocity needed to achieve maximum shear rates.

Determining the compatibility of various media for use in multimedium filters can be complicated. Intuitively it may be obvious that a separation will be

TABLE 8. - Fluidization results

d_m	Ar	Re_l	Re_{opt}	ϵ_{opt}	n
Sand:					
1.54-mm.....	53,530	306	57.6	0.56	2.885
1.00-mm.....	15,497	178	26.9	.56	3.220
.71-mm.....	5,486	79.4	14.2	.57	3.051
.50-mm.....	4,727	50.4	13.0	.69	3.607
MgO:					
1.54-mm.....	98,460	457	83.4	.65	3.899
1.30-mm.....	58,690	332	61.0	.65	3.875
1.00-mm.....	25,816	210	36.8	.65	4.030
.50-mm.....	3,218	77.7	10.3	.67	5.057
.42-mm.....	1,908	38.2	7.44	.74	5.421
.25-mm.....	402.5	11.0	2.86	.79	5.645
Olivine:					
0.50-mm.....	4,727	72.5	13.0	.74	5.676
.35-mm.....	1,667	38.7	6.85	.82	8.098
.25-mm.....	562	16.7	3.50	.83	8.751
Sand (10):					
0.84-mm.....	7,935	95.7	17.9	.65	3.873
.65-mm.....	4,524	63.4	12.6	.67	4.041
.42-mm.....	1,184	27.1	5.55	.68	4.162
Ballotini (10):					
2.38-mm.....	111,332	288	90.4	.59	2.230
1.30-mm.....	25,659	134	36.7	.59	2.429
.92-mm.....	7,313	69.9	17.0	.58	2.596
.78-mm.....	5,221	56.1	13.8	.58	2.611
Anthracite (10):					
2.59-mm.....	35,414	174.4	44.7	.66	3.242
2.38-mm.....	21,418	133.7	32.8	.66	3.417
1.84-mm.....	14,771	107.6	26.1	.68	3.719
1.30-mm.....	4,948	60.2	13.4	.67	3.795
Polystyrene (10):					
3.67-mm.....	18,920	116.8	30.4	.65	3.117
3.08-mm.....	14,724	102.9	26.1	.65	3.180
2.59-mm.....	10,483	84.5	21.2	.66	3.301
2.18-mm.....	7,409	69.1	17.1	.66	3.420

possible only if Re_l of the upper medium is smaller than that of the lower layer. Re_l also increases as a function of Ar, so it should be possible to predict separation based on the size of Ar. In practice, however, this becomes complicated by differences in particle shapes and the fact that some intermixing of layers may be desirable. For conventional filtration systems that operate with a downward flow, it is desirable to have the coarsest medium in the uppermost layer to achieve any benefit over single-medium filters. From this standpoint, a combination of MgO and sand is not a practical dual-medium filter.

Attempts to backwash a filter with the upper layer of 0.71-mm sand and the lower layer of 0.5-mm MgO resulted in almost total intermixing of both layers. The corresponding values of Ar and Re_l are very close (table 8). To get adequate differences in Re_l would require either substantially finer MgO or at best equivalently sized sand as the upper layer, neither of which is desirable.

MgO worked well with anthracite. Dual-medium filters of 0.71-mm MgO and 1.3-mm anthracite were easily restratified by backwashing. A sharp interface was produced between the two media after 15 min of backwashing at flow velocities of

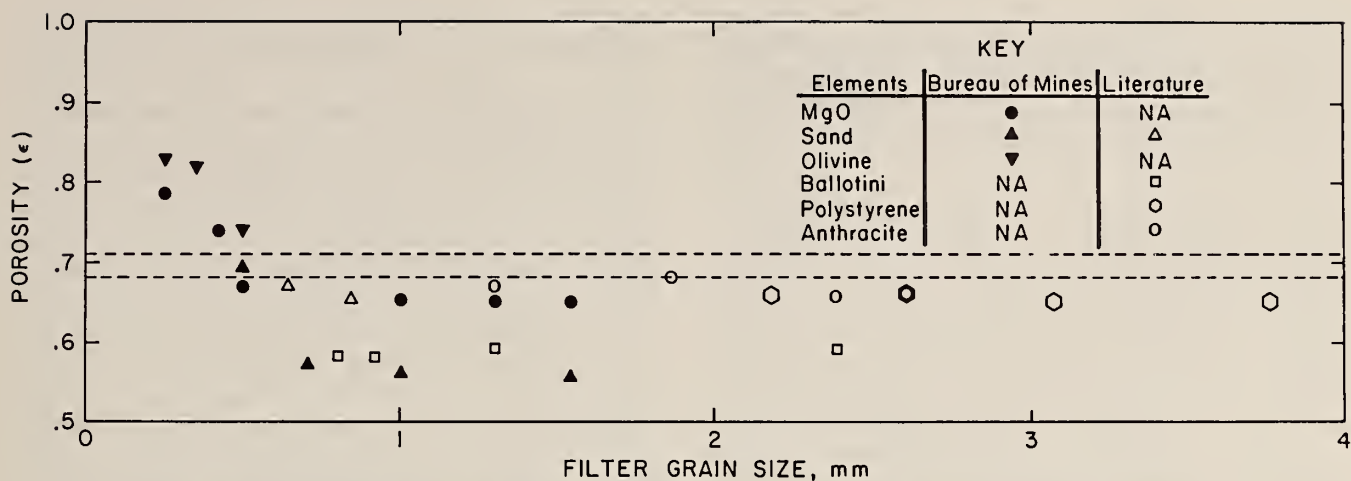


FIGURE 6.—Optimum bed porosities calculated from optimum Reynolds number. NA = Not available.

around 4 cm/s. Decreasing the velocity or duration of the backwash resulted in more intermixing of the two layers. An equivalently sized sand-anthracite filter was adequately but not so cleanly separated as the MgO-anthracite filter.

ATTRITION OF MgO FILTERS DURING BACKWASHING

Backwashing with air scour creates a significant amount of interparticle abrasion in the filter medium. Although this is an effective backwashing technique, it does increase attrition of the filter medium. The rubbing action of the grains of filter medium produces fines that can clog the bed in subsequent filtrations and also reduce the effective size of the medium. High attrition rates can result in noticeable changes in filter performance; head-loss rates can increase enough to drastically reduce production of filtered water. Angular particles such as anthracite or MgO would be expected to experience greater attrition than a spherical filter medium like sand.

Forty-eight filtration cycles were made on a 61-cm bed of 0.71-mm MgO to test the resistance of the MgO filter grains to attrition. A standard suspension of 25 ppm kaolin and 25 ppm milled sand, flocculated with 10 ppm alum, and a filtration velocity of 0.21 cm/s were used throughout. Each cycle consisted of a 6-h filtration run followed by 1 h of

backwashing. The backwashing procedure consisted of a few minutes of fluidization, followed by about 30 min of air-water scouring and then by 30 min of high-rate backwashing at 50-pct bed expansion. This latter step was employed to remove fines and dislodge air bubbles. This backwashing procedure was actually more intense and longer in duration than that normally used in commercial practice, since an air scour lasting 3 to 5 min and a 10- to 15-min fluidization at 20- to 50-pct bed expansion is usually sufficient to clean the bed (9). Total elapsed time for the 48 cycles to be completed was about 6 months. The MgO filter grains were kept under water in the filter column between cycles.

Data from the initial, final, and every fifth intermediate run are listed in table 9. Cycle 39 was used in place of

TABLE 9. - Durability study

Cycle	τ/τ_0	V/H	$F(\times 10^5)$	SC, g/cm ³
1.....	0.048	190	26	9.0
5.....	.081	440	19	20
10.....	.081	590	13	27
15.....	.042	640	6.5	31
20.....	.038	440	8.8	21
25.....	.046	440	11	21
30.....	.042	341	12	16
35.....	.042	296	13	14
39.....	.042	281	15	13
45.....	.040	326	12	16
48.....	.031	297	11	14

cycle 40 because the latter had an unusually large increase in head loss owing to air blinding of the filter. Variations in head loss rate account for most of the variation in the values of the filtration indexes.

The solid-capture index, SC, is plotted versus the number of cycles in figure 7. Filter performance initially improves, then decreases gradually until a fairly stable value of SC is reached in the vicinity of the 40th cycle. Experimental uncertainty exists because low pressures and small pressure differences are difficult to measure. Small pressure drops (head loss) for a bed result in high values of SC. The uncertainty in measuring these small pressure drops leads to the large error brackets for SC values. In spite of the large uncertainty bracketing each point, the trend shown in figure 7 appears to have statistical significance, and the value of SC near the end of the test has stabilized around 12 to 15 g/cm³. The behavior of the filterability index [$F(\times 10^5)$] essentially mirrors that of SC, and a stable level between 10 and 15 is reached about halfway into the test. Values for F and SC compare quite favorably with those given in table 3 for filtration of kaolin and milled sand.

Forty-eight backwash cycles is equivalent to a few weeks to several months of

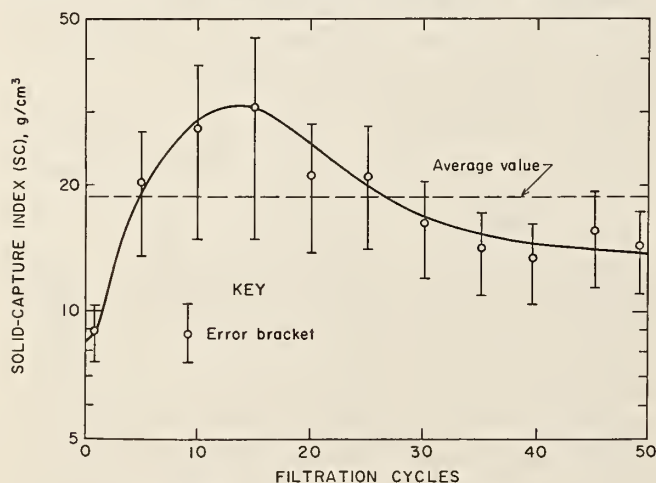


FIGURE 7.—Binding of MgO filters.

operation, depending on the application. The intensity and duration of the air scour, which is largely responsible for filter grain attrition, is equivalent to a period of normal commercial operation perhaps 6 to 10 times longer. Since filtration performance of the MgO has stabilized at an acceptable value during these tests, attrition of MgO does not seem to be a significant problem. Media losses to attrition and entrainment in conventional media can be as high as 5 to 15 wt pct in the first year of service (lower value for sand and higher value for anthracite). MgO losses are within this range; no significant decrease in bed depth was evident at the end of this test.

MgO durability was also evaluated by a standard friability test used to evaluate filter media (13). Friability is determined by calculating the fraction of sample by weight that remains larger than the effective grain diameter (d_{eff}) after milling. Samples are milled by steel balls in a metal cylinder which is tumbled end over end at 25 r/min for 15- and 30-min intervals. The milled samples are sieved, and the weight in each size fraction is compared to initial weights in those size fractions. Losses of 6 to 10 wt pct or less at 15-min milling and 15 to 20 wt pct or less at 30-min milling of the filter material larger than d_{eff} indicate that the filter material has good durability. Attrition losses of 10 to 15 wt pct and 15 to 25 wt pct for the two time intervals are tolerable for conventional filter media. A candidate filter material is rejected if losses are >20 wt pct or >35 wt pct, respectively, for the two time intervals (13). MgO passed the standard friability test easily. A sample with an original d_{eff} of 0.61 mm had losses of 4.0 wt pct and 4.5 wt pct for the 15- and 30-min milling. A second sample with d_{eff} equal to 0.47 mm showed losses of 5.5 and 7.0 wt pct for the two intervals. Both tests demonstrated very good durability for MgO.

POISONING OF MgO FILTERS BY HEAVY METALS

Dissolved heavy metals present in the influent will encounter an increase in pH and may precipitate as they pass through an MgO filter. Previous experience indicated that these precipitates adhere quite strongly to MgO granules. This could spoil the desirable surface properties of the MgO if the metal deposits are not removed by routine backwashing. If this were the case, it would become necessary to chemically strip the metals, most likely by adding either dilute acids or chelating agents at some stage in the backwashing.

In this study a 46-cm bed of 0.5-mm MgO was tested using 25 ppm suspensions of kaolin, both with and without dissolved heavy metals being present. Influent pH was adjusted to 7.0 ± 0.1 for all tests. The filter was put through a series of tests in which alternate filtrations were spiked with 5.0 ppm each of Cd^{2+} , Mn^{2+} , Ni^{2+} , and Zn^{2+} , all of which are

soluble at neutral pH at this level of concentration. Between tests the filter was backwashed with air-water scouring, followed by fluidization at high flow velocity. Filtration velocity was 0.34 cm/s.

Plots of filtrate turbidity, pH, and pressure for these tests are plotted versus time in figure 8. In the first test cycle no metals were added and filtrate pH remained fairly constant at 10.2 to 10.3 for the 6-h test cycle. Head loss was small, and an average of 90 pct of the turbidity was removed. In the second test cycle heavy metals were in the influent. Turbidity was decreased by 99 pct, but head loss increased dramatically. Coating of the MgO with precipitated metal hydroxides is indicated during this cycle by the steadily declining filtrate pH. Subsequent filtration without metals in the third test cycle showed a slight increase in the removal of turbidity and slightly lower filtrate pH than in the first test cycle. In the

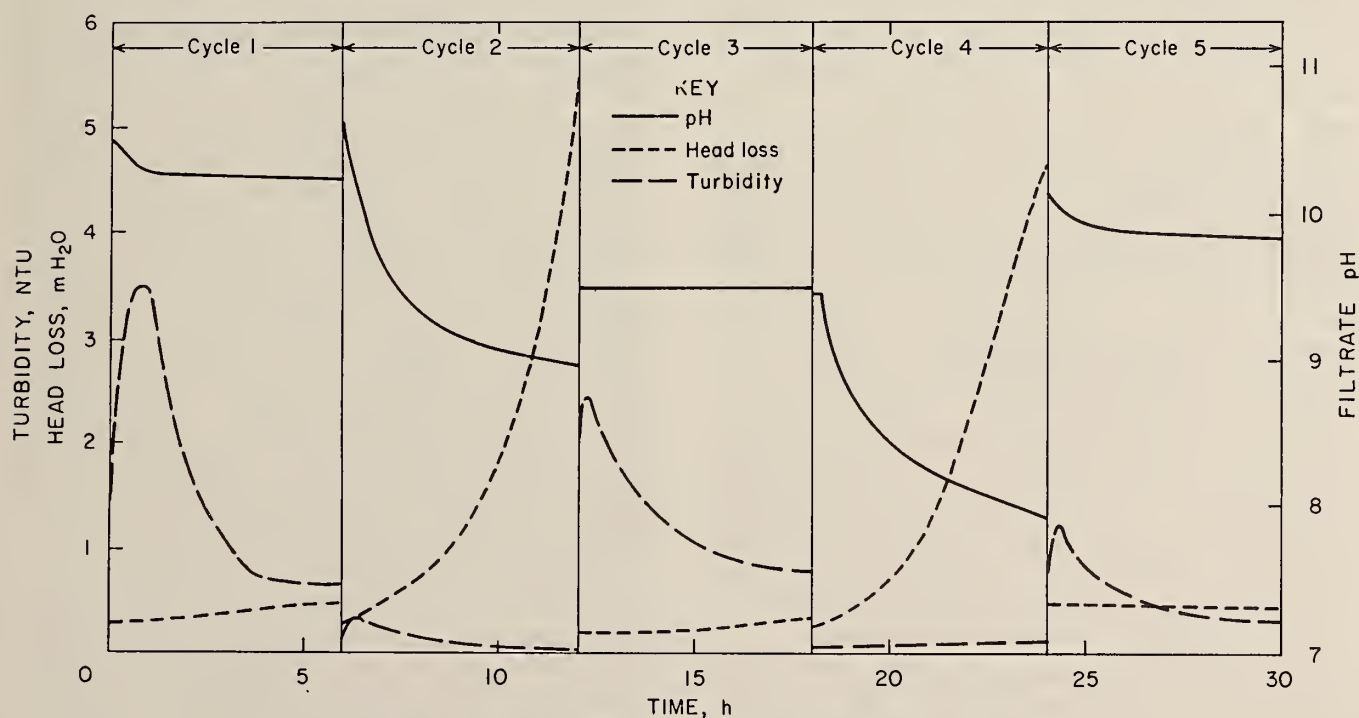


FIGURE 8.—Effects of dissolved heavy metals on MgO filtration.

fourth test cycle the dissolved metals were passed through the filter without the presence of kaolin. In this cycle the pH declined until it was only one unit higher than that of the influent, indicating that the reactive surface area of the MgO was virtually saturated with deposited precipitates of metal hydroxides. Head loss was large in this cycle, indicating that the precipitated metal hydroxides rather than flocculation of the kaolin were responsible for the increased resistance to flow observed in the second test cycle. In previous tests, flocculation was not observed in the reservoir containing the dissolved metals and kaolin suspension. In the fifth test cycle solid capture was again improved over that in the initial test; in fact, performance as measured by

filtration indexes actually improved steadily with increased exposure to heavy metals. Values of $F(\times 10^5)$ for the first, third, and fifth runs were 28, 14, and 3.0, and corresponding values for SC were 7.8, 12, and 24 g/cm³. For the second test cycle, which had both metals and kaolin present, $F(\times 10^5)$ was 60 and SC was 3.5 g/cm³. It is concluded that heavy metals present in water to be filtered will be precipitated as hydroxides and will be bound to the MgO filter grains. These precipitated hydroxides are only partially removed by backwashing; however, their presence on the MgO filter grains does not reduce the filterability of the MgO. Too high a level of heavy metals in the turbid water to be filtered will result in accelerated increases in head loss.

FILTRATION PARAMETERS

EFFECT OF SURFACE CHARGE AND PARTICLE SHAPE ON FILTRATION

It has never been fully demonstrated to what effect the positive surface charge of the MgO contributes to its ability to remove particulate, although this has been previously suggested as a primary factor (2-3). In early contact filtration test studies, which used fine-mesh active MgO, surface charge interactions undoubtedly contributed greatly to the filtration of unflocculated asbestos fibers. But in applying this concept to practical deep-bed filtration, both surface area and activity were greatly reduced in the transition. Deeper beds of coarse, dead-burned (fused) periclase were used instead of active (porous) MgO; furthermore, flocculation with aluminum salts was necessary to achieve efficient particulate removal. This would all tend to minimize any surface charge effect.

The ultimate removal efficiency of a sand filter is equal to and sometimes slightly better than that of MgO, but the amount of material that can be collected (run length) of the MgO is often double that of sand. This can be partially explained by the greater porosity of the MgO; more material can be deposited before critical shear stresses are reached.

Increased porosity means increased average pore diameter, as would be the case in using a coarser filter medium. Capture efficiency, however, usually decreases as pore diameter increases. Some extra compensating mechanism allows the MgO to achieve removal rates nearly equivalent to those for sand. Surface charge and particle shape (altered hydrodynamics) are two possible mechanisms.

A factorial study was performed to qualitatively evaluate the effect of surface charge and grain shape on particle removal. Four combinations of medium shape and charge were tested using a standard pH 7.0 suspension of 50 ppm kaolin with no flocculant. Shallow beds (15.2 cm) of 0.5-mm filter media were tested during 2-h runs. The size of the media tested is toward the small end of the sizes used in deep-bed filters.

The four filter media were MgO (+, angular), sand (-, spherical), quartzite (-, angular), and MgO-coated sand (+, spherical). Zeta potentials were determined for crushed samples of media suspended in distilled water. The zeta potential of kaolin in distilled water was also measured and found to be -27 ± 5 mV. First-order removal coefficients were calculated from the ratio of influent and filtrate turbidities:

TABLE 10. - Effects of filter medium surface charge and particle shape on kaolin removal

Medium	Zeta potential, mV	No. of trials	Removal coefficient, m^{-1}	
			λ_o	λ_f
Angular shape:				
MgO.....	20±7	4	7.5±0.6	6.5±0.4
Quartzite.....	-28±3	3	5.4±1.1	11 ±1.5
Spherical shape:				
MgO-coated sand	(+)	13	7.9±1.8	9.2±1.9
Sand.....	-15±6	4	3.0± .6	8.4±1.3

¹The last trial was deleted because MgO coating was being lost. Values for λ were much lower than for the other 3 trials.

$$\lambda = \frac{\ln (\tau/\tau_o)}{l} \quad (10)$$

Results are given in table 10. The clean-bed coefficient (λ_o) is based on the filtrate turbidities taken at 1 and 15 min into the test, while the final turbidity (2 h) is used to calculate λ_f .

In the initial stages of filtration there was little deposited particulate, so particle-to-medium interactions should be at their maximum. Here MgO, whether as a spherical or an angular grain, is the most efficient filter by a substantial margin. The negatively charged media, whether angular or spherical, had lower removal efficiencies, but the magnitude of the (-) zeta potentials makes less difference than the shape of the medium; the more negative and angular quartzite is more effective than spherical sand. The negative media exhibited increasing filtration efficiency with time; values of λ_f are approximately double those of λ_o . The MgO efficiency declines very slightly in this time; the MgO-coated sand efficiency increases, but with less than experimental uncertainty. Negatively charged media actually become the more efficient filters as deposits form in the filter owing to enhanced hydrodynamic and mechanical interception mechanisms. Surface charge interactions were more important than shape for enhancing clean-bed-particle collection. It is also of interest that head loss rates were lower for the positively charged media, averaging 3.5, 0, 5.3, and 8.4 cm H₂O for the MgO, MgO-sand, sand, and quartzite, respectively. Unfortunately, the experimental uncertainty in

pressure readings is at least 3.5 cm H₂O, but it would seem that particle shape (hence bed porosity) alone is not totally responsible for observed head loss behavior. It may be that different modes of particle deposition can significantly alter the development of head loss. Most deep-bed filters have coarser media, and flocculant is used to neutralize particle charge and increase particle size, so the surface charge of the medium is probably of little practical importance in operating conventional deep-bed filters.

REMOVAL AND HEAD LOSS COEFFICIENTS

Empirical models have been used to describe the head loss and filtrate quality as functions of time and filter-bed depth. The clean bed head loss can be calculated from the Kozeny-Carmen equation (5):

$$\left. \frac{\delta H}{\delta l} \right|_0 = \frac{5vV(1-\epsilon)^2}{g\epsilon^3} \left(\frac{6}{d_m} \right)^2, \quad (11)$$

where H = head loss,

l = the filter depth,

v = the kinetic viscosity,

g = the gravitational acceleration,

ϵ = the bed porosity,

v = the filtration velocity,

and d_m = the grain size of the filter bed.

A good filtration run exhibits a nearly linear increase in head loss with respect to time, which is closely approximated by

$$H = H_i + k_H v C_o t, \quad (12)$$

where H = head loss,

H_i = initial head loss,

C_o = the influent concentration of suspended solids,

t = the elapsed time,

and k_H = the head loss coefficient (cm^3/g).

This assumes particle removal rates of >99 pct. The head-loss coefficient is very similar to the SC index of equation 2. The concentration profile with respect to depth has been modeled as a first-order process:

$$\frac{\delta C}{\delta l} = -\lambda C, \quad (13)$$

where λ is the removal coefficient (5). Concentration decreases logarithmically with depth as given in equation 10. A variety of other more complicated models are available for detailed analysis of deep-bed filters (6-7). To date, it has not been possible to predict these engineering parameters, which are useful in filter design, and thus they must be obtained, at least in part, by experiment (5-6).

Beds of 0.5-mm sand and MgO were tested against 12.5-ppm kaolin suspensions at pH 7.0 ± 0.5 . Flocculant was not used. Bed depths were varied between approximately 5 and 50 cm. A water manometer was used rather than the usual pressure gauges to improve sensitivity and accuracy. (Pressure could be measured to ± 2 mm H_2O .) For shallow beds, a known weight of filter medium was used rather than attempting to fill filter columns to a particular depth, since variations in packing lead to relatively large changes in bed

depth. Turbidity was used to calculate removal rate, and clean-bed-filter coefficients were calculated from the first two turbidity readings. Results are presented in table 11.

Attempts to fit $\ln(\tau/\tau_o)$ as a linear function of depth were unsuccessful; although removal decreased with increasing depth, there was considerable random scatter around any type of linear plot. For this reason, values of λ_o were merely tabulated and an average value given. Since turbidity removal was less than 99 pct, SC was calculated rather than k_H for head loss rate data.

In general, the MgO filter gave slightly better removal rates and slightly lower head loss rates than the sand filter. Results varied widely; the standard deviation of duplicate tests is 20 to 50 pct of mean values. It appears that SC values improve for sand with increased bed depth, while values for MgO fluctuate randomly. Measurement uncertainties for turbidity and pressure are much smaller than the variation in results. Preparation of suspensions is also carefully controlled, so it would seem that a fair amount of randomness is intrinsic to the process of deep-bed filtration. Packing irregularities may also be partially responsible for the variability of results. Although these results are not quantitatively precise, it can be seen that MgO does perform slightly better than sand without the intervening variable of

TABLE 11. - Head loss and removal coefficients of MgO and sand filters for the filtration of kaolin

Bed depth	λ_o , cm^{-1}	F ($\times 10^5$)	SC, g/cm^3
MgO:			
9.2 cm.	0.003	44	6.3
18 cm..	.048	18	18
36 cm..	¹ .024	¹ 34	¹ 9.1
46 cm..	.035	29	7.8
Mean..	.035 $\pm$ 0.18	36 $\pm$ 10.5	9.4 $\pm$ 4.6
Sand:			
10 cm..	.017	50	3.3
20 cm..	¹ .014	¹ 56	¹ 4.4
40 cm..	.028	34	7.5
Mean..	.018 $\pm$ 0.005	49 $\pm$ 8.3	4.9 $\pm$ 1.5

¹Average of 2 test values.

flocculation. These results are in qualitative agreement with the bulk of the experimental data presented in this paper.

pH AND CHEMICAL EFFECTS

The major filtration characteristic of MgO, other than its positive surface charge, is its basicity. Unbuffered water passing over a bed of MgO will experience an increase in pH of several units. At filtration velocities of 0.2 to 0.3 cm/s, water entering a typical bed of periclase (inert MgO) at neutral pH will exit at pH 10 to 10.5. Salts of Al^{3+} and Fe^{3+} are amphoteric; consequently, a shift in pH could be expected to significantly alter their solubility. Both Al^{3+} and Fe^{3+} salts are common coagulants and/or flocculants used in water treatment to destabilize colloidal dispersions by neutralizing surface charge. Often relatively large quantities of Al^{3+} or Fe^{3+} are added to produce $\text{Al}(\text{OH})_3$ or $\text{Fe}(\text{OH})_3$ flocs that further clarify water by enmeshing particulates. Alum flocculation was used in the majority of the filtration tests, and while this would tend to decrease favorable surface-charge interactions between particulates and MgO, the effect of pH could conceivably be important. Effects of pH generally were not investigated during previous filtration tests.

A series of tests were run with 24-ppm kaolin suspensions in the presence of either 15 ppm Al^{3+} as alum or 30 ppm Fe^{3+} as FeCl_3 . This concentration of Al^{3+} is about 5 to 20 times greater than those used in previous tests. (Concentrations

were previously reported as ppm alum; formula weight is 474.) Influent pH was varied between 4.5 and 7.0, and filtration velocities were 0.33 to 0.35 cm/s. Results are given in table 12. A strong pH effect is evident for the Al^{3+} -treated suspension, as turbidity removal is observed to be much poorer at higher influent pH. At pH 6.9, removal is so poor that head loss is almost negligible; hence value of F is low. Fe^{3+} removal is poor regardless of pH. Filtrate pH decreases considerably throughout the test if alum is present, but changes very little with FeCl_3 .

The differences in filtration results are probably due to the differences in solubility of the two salts. Al^{3+} readily forms soluble hydroxy complexes, and the region of minimum solubility where $\text{Al}(\text{OH})_3$ formation predominates spans a fairly narrow pH range of 2 pH units centered around pH 5. The solubility diagram of Fe^{3+} is somewhat similar, though $\text{Fe}(\text{OH})_3$ remains more insoluble over a much wider pH range; soluble hydroxy complexes are not formed appreciably between pH 3.5 and 13. Typical applications in water treatment use Al^{3+} at pH 5 to 9 and Fe^{3+} at pH 3 to 8 (14).

Apparently, efficient removal of the kaolin-Al suspension requires a certain amount of soluble Al to interact with the MgO surface. Sufficient soluble Fe is not available at these pH levels; consequently, bonding is poor and little kaolin-Fe suspension is retained. MgO is not likely to be used to collect Fe^{3+} -flocculated suspensions in practice because dissolution of the MgO will

TABLE 12. - Effect of influent pH on filtration of kaolin suspension flocculated with Al^{3+} and Fe^{3+} salts

t_o , NTU	τ/τ_o	v , cm/s	V , L/cm ²	V/H	F ($\times 10^5$)	Influent pH
WITH 15 ppm Al						
35.....	0.043	0.340	3.67	32.6	130	4.5
23.....	.347	.347	2.50	1,010	34	6.9
27.....	.180	.330	4.75	25.5	700	5.6
28.5.....	.012	.349	5.02	14.9	82	4.5
WITH 30 ppm Fe						
25.....	0.221	0.353	1.27	36	610	6.7
29.....	.370	.350	1.26	60	620	4.6

probably be excessive at pH levels much below 5.0.

The mechanisms of these interactions are complex and further complicated by the fact that pH levels at the MgO surface are likely to be considerably higher than those of the bulk solution. The effect of pH on particle- and medium-surface charge would also have to be considered. What effect pH had on filtration tests with much smaller amounts of the Al is unclear, but it is probably at least as important as surface charge effects.

SCALE FORMATION AND MUDBALLING

Another consequence of the basicity of the MgO is scale formation. Water with appreciable calcium hardness and carbonate or phosphate alkalinity will form insoluble calcium compounds due to the increase in pH. Cementation of MgO filters was observed in tests with several water samples. During backwashing, the medium tended to lift as a plug rather than fluidizing. Usually, air scour was enough to break up the scale and clean the filter. Formation of scale did not seem to seriously impair removal rates or cause excessively high head-loss rates during most filtration runs.

Mudballing is one problem that can result from scale formation in MgO filters. In conventional filter media, mudballing results from the compression of flocculated solids into a cake at the surface of the filter. The relatively large chunks of cake are not entrained in backwash at practical fluidization rates. Depending on their size and density, these will either stay at the surface or sink to the bottom of the bed. In the field tests with the process water used for cutting granite, mudballing manifested itself differently. Cementation of the MgO was evident in the lower part of the filter, and although the bed was fluidized for the most part during backwashing, a rim of the cemented material remained along the outer edge at the bottom of the filter. This probably would be avoided if backwash air and water were better distributed at the bottom of the filter. However, control of

scale-forming tendencies in the process water is recommended.

Another type of cementation was observed in later tests at the same plant. Chunks of cemented MgO were again observed at the bottom of the filter bed. These were removed for inspection and are shown in figure 9. Their most interesting feature is the presence of grains of anthracite adhering to one side of the cemented chunks. Evidently these chunks formed at the MgO-anthracite interface and worked their way to the bottom of the filter during backwashing. These were probably formed as a result of compression of deposited solids at the



FIGURE 9.—Cemented MgO chunks containing anthracite grains.

MgO-anthracite interface, but scale formation may also be responsible. Sand filters and sand-anthracite filters

apparently were not affected in these tests.

SUMMARY AND CONCLUSIONS

Under the right circumstances MgO filters offer significant advantages over similar conventional sand filters. Bureau results suggests that 0.5- to 0.71-mm MgO will filter water pretreated with alum better than equivalently sized particles of conventional filter sand. Particulate removal was approximately equal for the two media under these conditions, but much more water could be passed through the MgO filters before breakthrough. The porosity of the MgO filter bed is about 1.3 times greater than that of the sand filter bed, which probably accounts for its smaller rate of head loss and large run lengths before breakthrough.

Current trends in conventional water treatment include the use of coarser media in conjunction with polymer flocculants to increase solid-loading capacity. The flocs created by polymer addition are more resistant to higher shear rates than alum flocs; consequently, higher filtration velocities are used. Excessive head loss rather than high turbidity tends to limit run length. Under these conditions MgO apparently offers little advantage over sand other than a slight reduction in head loss rate.

In tests of recycled process water, elevated levels of dissolved solids usually were found. The MgO filters were tolerant to moderate levels of calcium hardness and carbonate alkalinity, provided adequate backwashing with air scour was available.

No single solid removal mechanism can be definitively identified as the one responsible for the improved filtration observed with granular MgO. In contact filtration of asbestos fibers, surface charge effects were almost certainly predominant, but in shifting to granular MgO filters the specific surface area was greatly reduced and alum flocculation was necessary to achieve efficient solid removal. Both factors tend to indicate decreased importance of surface charge

effects in comparison with mechanical effects. pH-chemical effects may also be important.

Granular MgO (periclase) possesses the necessary durability to be a filter medium; no drastic attrition effects were noticed in field tests or in laboratory longevity studies. MgO is also compatible with anthracite as a dual-medium filter, whereas sand-MgO filters are not likely to stratify in a workable manner except possibly in upflow filters. Bed poisoning by dissolved metals apparently is not a problem.

Filtration is just one step in the overall water-treatment process. Optimization of the clarification process will most likely outweigh optimization of the filtration process, since the former removes by far the larger amount of solids. In instances where mine and mineral-processing water is only moderately contaminated by suspended solids, direct filtration of the mine water without clarification may be an attractive alternative. A filter that can effectively remove suspended solids without pretreatment would be desirable. Comparison of filtration tests where 0.5-mm MgO is used to filter untreated kaolin and 0.71-mm MgO is used to filter alum-treated kaolin are encouraging. Solid capture indexes (SC) and filterability indexes (F) were better by almost an order of magnitude when no flocculant was added. Flocculant evidently adds considerable bulk to the suspended solids and contributes heavily to head loss. Reduced head loss is potentially the most beneficial advantage in employing MgO filtration. Rather than concentrating on comparing MgO and other media as conventional water filters, future research could explore the use of novel materials in novel filtration methods. Various grades of MgO with intermediate activity and hardness should be evaluated as filter media to determine whether the surface properties of MgO can be better utilized.

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APPENDIX.--NOMENCLATURE

SYMBOLS

a_p	radius of suspended particle, mm
A_r	ratio of buoyant to viscous drag forces acting on a particle, dimensionless
C	suspended solids concentration in filtrate, mg/L
C_o	influent suspended solids concentration, mg/L
d_{eff}	effective grain diameter, mm
d_m	diameter of the filter medium granules, mm
d_p	diameter of particulate being filtered, mm
e	charge on an electron, eV
F	filterability index, dimensionless
f	subscript denoting fluid
g	acceleration due to gravity, 9.8 m/s^2
H	head loss, cm H_2O
h	constant in London group, N_{L0} ; dimensionless
h_b	bed depth, m
h_{bo}	bed depth of unfluidized bed, m
H_i	initial head loss, cm H_2O
k	Boltzman's constant, cal/(mol·deg)
k_H	loss coefficient, cm^3/g
l	filter depth, m
M	represents metal
m	medium property
m_j	molality of jth ion
N_{DL}	double-layer filtration parameter
N_{E1}	electronic group 1 filtration parameter
N_{E2}	electronic group 2 filtration parameter
N_{L0}	London group filtration parameter

SYMBOLS--Continued

N_R	relative size group filtration parameter
N_G	gravity group filtration parameter
n	parameter in Reynolds number determinations, dimensionless
P	element symbol for phosphorus
P_e	Peclet number
p	particle property
R	ideal gas law constant, 1.987 cal/(mol·deg)
Re	Reynolds number, dimensionless
Re_i	Reynolds number corresponding to unhindered settling of filter grains, dimensionless
Re_{opt}	optimum Reynolds number, dimensionless
SC	solid capture index, g/cm ³
$S.G.$	specific gravity, g/cm ³
T	absolute temperature, K
t	time, h or s
V	filtrate volume per unit area, L/cm ²
v	filtration velocity, cm/s
z_j	valence of jth ion

GREEK LETTERS

δ	differential notation	ν	kinematic viscosity, μ/p or m ² /s
ϵ	porosity, dimensionless	π	constant, 3.1416
ϵ_0	porosity of unfluidized bed, dimensionless	ρ_f	density of fluid, kg/m ³ or g/cm ³
ϵ_{opt}	optimum porosity, dimensionless	ρ_p	density of particulate, kg/m ³ or g/cm ³
ϵ_d	dielectric constant of fluid, dimensionless	τ	turbidity, NTU
κ	inverse double-layer thickness, m ⁻¹	τ_0	initial turbidity, NTU
λ	removal coefficient, cm ⁻¹	Ψ_m	surface potential of filter medium, mV
μ	viscosity, kg/(m·s)	Ψ_p	surface potential of particulate, mV

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